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Final Report
Contract CPA 22-69-78

FEASIBILITY STUDY OF NEW SULFUR OXIDE CONTROL PROCESSES FOR APPLICATION TO SMELTERS AND POWER PLANTS

Part III: The Monsanto Cat-Ox Process for Application to Power Plant Flue Gases

Prepared for:

U.S. DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
DURHAM, NORTH CAROLINA

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Prepared for:

U.S. DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
DURHAM, NORTH CAROLINA

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CONTENTS

FOREWORD v

I INTRODUCTION 1

II OBJECTIVES 5

III SUMMARY 7

IV PROCEDURES 15

A. Formulation of Models 15

B. Cost Factors 16

C. Preparation of Technical Data and Cost Estimates . . 16

V PROCESS DESCRIPTIONS 17

A. Integrated Cat-Ox System for Power Plants 18

B. Cat-Ox System for Application to Existing Power Plant 25

VI PROCESS DATA AND COST ESTIMATES 29

A. Material Flows 29

B. Capital Cost Estimates 34

C. Operating Cost Estimates 38

VII GENERAL DISCUSSION 39

A. Evaluation of the Cat-Ox System 39

B. By-Product Values 42

C. Variation of Bases of Cost Estimates 44

1. Load Factor 44

2. Armortization Period 45

3. Fixed Charges 46

REFERENCES 47

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A block flow diagram was prepared.
Mass and volume flow balances were estimated. Engineering estimates of the capital investment and the total annual cost (including both fixed and variable charges) were prepared. These engineering estimates were applied to previously created model gas streams and the total annual cost for each was determined. From these annual costs, incremental production costs were estimated. The technical constraints on the application and operation were qualitatively appraised; the economic constraints were appraised (quantitatively, to the extent permitted by the data). The current state of development was assessed, identifying any technological deficiencies whose elimination might enhance applicability.

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CONTENTS (Concluded)

APPENDIXES

A. MODELS FOR HYPOTHETICAL POWER PLANTS A-1

B. COST FACTORS USED IN MODEL STUDIES B-1

C. STACKS FOR USE ON CONTROLLED SULFUR DIOXIDE EMISSION SOURCES C-1

D. ESTIMATION OF THE VALUES OF SULFUR BY-PRODUCTS D-1

ILLUSTRATIONS

Figure 1 Integrated Cat-Ox System for Power Plant Model B 19

Figure 2 Cat-Ox Reheat System for Power Plant Model A 20

TABLES

Table I Summary of Estimated Costs -- Cat-Ox Reheat System for Power Plant Model A 11

Table II Summary of Estimated Costs -- Integrated Cat-Ox System for Power Plant Model B 12

Table III Model A -- Gas Flows in Cat-Ox Reheat System 30

Table IV Model B -- Gas Flows in Integrated Cat-Ox System 31

Table V Model A -- Contaminant Removal in Cat-Ox Reheat System . 32

Table VI Model B -- Contaminant Removal in Integrated Cat-Ox System 32

Table VII Power Plant Model A -- Summary of Capital and Operating Costs for Cat-Ox Reheat System 34

Table VIII Power Plant Model B -- Capital Investment for Integrated Cat-Ox System 35

Table IX Power Plant Model B -- Summary of Capital and Operating Costs for Integrated Cat-Ox System 36

Table D-1 Estimated Sulfuric Acid Demand in Selected Producing Areas, 1966 . . . . . D-6

Table D-2 Phosphoric Acid Production Costs D-7

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FOREWORD

The final report for this study is presented in four separate and independent parts:

Part I: The Monsanto Cat-Ox Process for Application to Smelter Gases

Part II: The Wellman-Lord SO_2 Recovery Process for Application to Smelter Gases

Part III: The Monsanto Cat-Ox Process for Application to Power Plant Flue Gases

Part IV: The Wellman-Lord SO_2 Recovery Process for Application to Power Plant Flue Gases

Information for use in this study was supplied to Stanford Research Institute by Monsanto Company and Wellman-Lord, Inc. under terms of confidentiality agreements between the U.S. Department of Health, Education, and Welfare, Stanford Research Institute, and each of the cooperating companies. In accordance with the agreements, Monsanto Company and Wellman-Lord, Inc. have reviewed and released the parts of the report dealing with their respective processes. The rights of prior review and release are designed solely to permit the cooperating companies to assure themselves that no proprietary or confidential data are being revealed; they are not intended to restrict Stanford Research Institute's rights and responsibilities to report its conclusions so long as there is no incidental disclosure of confidential information. Accordingly, the release of the reports by Monsanto and Wellman-Lord does not imply that these companies necessarily concur in all or any of the opinions, judgments, or interpretations of fact expressed by the author, who assumes sole responsibility for the report content.

I INTRODUCTION

Under the Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry (Contract No. PH 86-68-85), Arthur G. McKee & Company and its subcontractor, Stanford Research Institute, carried out evaluations of a number of sulfur oxide control processes as they might be applied to offgases from nonferrous smelting. To permit evaluation of the technical and economic feasibility of these control processes, a number of models of smelters were created. Stanford Research Institute carried out the studies necessary to determine the availability of markets for sulfur by-products open to smelters in various areas, and the allowable production costs that the smelters would have to attain in order to break even on the sulfur recovery operations.

The Division of Process Control Engineering of the National Air Pollution Control Administration (DPCE-NAPCA) desires to extend the usefulness of the foregoing study by adding to it technical and economic evaluations of new and potentially promising sulfur oxide control processes. It also wishes to evaluate the same new processes for application to power plants. Completion of these preliminary evaluations of the processes will help determine their potential commercial acceptability.

DPCE-NAPCA has a specific interest in at least two control processes being offered commercially, the Monsanto Cat-Ox process and the Wellman-Lord SO_2 Recovery process. However, both processes are proprietary, and, as a matter of policy, DPCE-NAPCA does not wish to obtain proprietary and confidential information on the processes. It does, nevertheless, wish to obtain evaluations in nonconfidential terms. Broadly, DPCE-NAPCA wishes to obtain estimates of the capital and annual costs of the control systems for each of the assumed applications, together with appraisals of the technical constraints on each process and of the current states of development of the processes.

Stanford Research Institute was requested by DPCE-NAPCA to carry out evaluations of the processes under the terms of confidentiality agreements between the Department of Health, Education, and Welfare, the owners of the proprietary processes, and Stanford Research Institute. SRI, acting as a disinterested third party, was to make analyses of the processes using information obtained from Monsanto Company and Wellman-Lord, Inc., and to report the results to DPCE-NAPCA without compromising any of the Monsanto or Wellman-Lord confidential data.

Because of the requirements of confidentiality, it is not permissible to describe certain features of the Cat-Ox and Wellman-Lord processes. The corresponding portions of the systems have had to be represented only in terms of their general functions, and SRI's evaluation of these portions has had to be presented in the form of conclusions without supporting data or reasoning. In other instances, the parts of the systems could be described in general, but specific details and design parameters could not be revealed.

Within the scope of the present project, it would obviously have been impossible to inspect and evaluate independently all the company records and design data even had the cooperating companies been requested to permit this and had they acceded to the request. The author of this report, who also conducted the study, evaluated the information provided at his request, using his own knowledge and relevant data from the literature and other available sources. Whenever apparent discrepancies or uncertainties were noted in the information, efforts were made to secure verification or clarification from the companies. In instances where resolution of questions was not possible, or the information required proved to be simply unavailable, the author employed his best judgment.

Throughout the following sections of this report, information for which other sources are not specifically cited was generally obtained from the cooperating companies and accepted by the author either because it could be verified from other sources or because it appeared reasonable. In other instances, information or estimates were provided by the com-

panies that could not be verified independently or judged for reasonableness; in such cases, the companies have been specifically cited as the sources. In still other instances, the author did not accept the information or estimates provided and in some cases substituted his own; such cases have also been specifically noted.

The cooperating companies, at their own option and through substantial efforts, provided the basic capital cost estimates for the model control systems and the information for estimation of operating and maintenance costs. The author in this case acted as a reviewer rather than as an estimator. The estimates were checked for reasonableness and for possible errors or omissions. For some components and cost factors, the author modified the estimates, or substituted others of his own where he judged them to be more appropriate than those supplied to him. The author also prepared cost estimates for some auxiliary systems, using separate data sources.

By specification of the power plant and smelter models, and by review of the results, an effort was made to ensure that the cost estimates for both control systems were made on strictly comparable bases. Although it is unlikely that this objective has been met fully, the deviations are probably within the precision of the estimates themselves.

For convenience, and at the request of DPCE-NAPCA, this final report is presented in four separate and independent parts. This part deals only with the Monsanto Cat-Ox process as applied to control of sulfur oxides in the flue gases from power plants.

II OBJECTIVES

The objectives of the part of the study covered by this report are as follows:

1. To prepare block flow diagrams of the Cat-Ox system showing its configuration and its relation to the boiler plant in which the sulfur oxides are generated.

2. To present estimated mass and volume flow balances for the Cat-Ox system.

3. To prepare preliminary engineering estimates of the capital investment and the total annual cost (including both fixed and variable charges) for the Cat-Ox system. From the estimate of total annual cost, secondary estimates are to be made of the corresponding incremental cost of producing electricity, both on the gross basis (without allowance for by-product recovery credits) and on the net basis (with allowance for by-product recovery credits).

4. To make a qualitative appraisal of technical constraints on the application and operation of the control system.

5. To appraise (quantitatively, to the extent permitted by available data) the economic constraints on the application of the control system.

6. To assess the current state of development of the Cat-Ox system, identifying any technological deficiencies whose elimination might enhance the applicability of the system to power plants.

The accomplishment of the objectives is subject to any restrictions that may be imposed under the terms of the confidentiality agreement between the Government, Monsanto Company, and Stanford Research Institute.

III SUMMARY

The Monsanto Cat-Ox system for sulfur oxides recovery is essentially an adaptation of the well-known contact process for sulfuric acid manufacture. The gas containing both sulfur dioxide and oxygen is passed through a fixed bed of catalyst at an appropriate temperature and most of the sulfur dioxide is oxidized to sulfur trioxide. The gas is then passed through an absorption tower where the sulfur trioxide is absorbed in recirculated sulfuric acid. The Cat-Ox system has been developed for use on gas (primarily power plant flue gases) that contain dilute concentrations of sulfur dioxide -- typically 0.1 to 0.4 percent. It differs from the conventional contact process plant in three principal respects:

1. The feed gas entering the system either must be already at a temperature high enough for conversion of the sulfur dioxide to the trioxide in the catalytic converter, or else auxiliary heat must be supplied to raise the temperature. Because of the diluteness of the sulfur dioxide, the plant is not autothermal; that is, the heat released by the oxidation of the sulfur dioxide is insufficient to preheat the feed gas to the reaction temperature.

2. The system operates on wet gas. The feed gas is not dried before it enters the converter.

3. The heat in the exit gas is used to a greater or lesser degree to concentrate the sulfuric acid formed in the final absorption step.

For power plant applications the Cat-Ox process is offered in two forms, one for use in new plants and one for previously existing plants. The Cat-Ox system for new plants is constructed as an integral part of the boiler and steam generating system. It includes some of the normal components of a boiler plant, including the economizer, the electrostatic precipitator, and the air heater, but not all the costs of these components are properly chargeable against sulfur oxides recovery. The

Cat-Ox system for application to existing plants, which is termed a "Cat-Ox reheat system," is added to the power plant at the point where the flue gas is normally delivered to the stack, and the entire cost is chargeable to sulfur oxides recovery. The flue gas must be reheated to the reaction temperature by burning supplementary fuel, and it is necessary to provide a preheater to transfer the heat to the flue gas entering the converter and a heat exchanger subsequently to transfer part of the heat in the converter exit gas to the flue gas entering the system.

Because the gas is not dried before conversion of the sulfur dioxide to sulfur trioxide, the amount of sulfuric acid mist formed during the gas cooling and absorption steps is relatively much higher than in the conventional contact process. Hence, a high-efficiency mist collector must be used to recover the mist from the tail gas. In the Cat-Ox process, a fiber-bed mist eliminator is employed.

The concentration of the sulfuric acid produced by the system is 78 percent. In the integrated Cat-Ox system, stronger acid can be produced only if heat is diverted from the power plant cycle for use in concentrating the acid. A more complicated system and additional equipment are also required. In the Cat-Ox reheat system, stronger acid could be produced by adding the extra equipment and burning a greater amount of supplementary fuel. However, it should be possible to concentrate the acid much more economically with a separate, conventional acid concentrator.

All the basic concepts of the Cat-Ox process have been demonstrated previously, although technical problems remain with respect to specific applications of the process. The principal problems (actual or potential) in the specific Cat-Ox system for power plant applications are associated with fly ash. The fly ash must be removed from the flue gas with very high efficiency in a hot, dry process. Dust not removed from the entering flue gas tends to plug the catalyst bed in the converter, contaminate the product acid, and plug the fiber-bed mist eliminator. Monsanto uses an electrostatic precipitator (or a combination of cyclone separators followed by a precipitator) for the hot, dry gas cleaning operation. The rate at which the catalyst bed plugs depends, of course, upon the quantity of dust

passing through the precipitator. When the catalyst bed becomes plugged, it is necessary to remove the catalyst for cleaning. The cleaning operation requires a shutdown of two to three days and involves the loss of about 2.5 percent of the catalyst by attrition and breakage (Monsanto estimates). Monsanto estimates that it will be necessary to clean the catalyst at three-month intervals if the dust passing to the catalyst bed does not exceed the quantity allowed for in the Monsanto design. The precise quantity of dust permissible in the Cat-Ox system is considered confidential by Monsanto. However, the required dust collector efficiency on fly ash from firing of pulverized coal is greater than 99.6 percent, which is higher than is now being provided commercially in fly ash precipitators.

Of the dust that passes through the converter, a portion is collected in the acid in the absorption tower, and the remainder is collected with very high efficiency by the fiber-bed mist eliminator, where the buildup of solids produces an increase in the resistance to gas flow. Monsanto reports that it has developed a mist eliminator that can be washed periodically, while on stream, to remove the accumulated solids. However, they declined to furnish information on its design and operation; the author is therefore unable to judge what the actual status of this development may be, but questions whether there has yet been an adequate demonstration of the ability to keep the mist eliminator free of solids buildup over long periods of continuous operation.

The amount of maintenance required for the Cat-Ox system will be acutely dependent on the performance of the electrostatic precipitator, even in the absence of outright precipitator failure. Even a small deterioration of the precipitator performance may increase the quantity of dust entering the converter by a large factor, with a corresponding reduction in the length of the intervals between shutdowns for cleaning. It will therefore be necessary to attain high standards of maintenance and operation of the precipitators -- probably much higher than the standards normally reached in the U.S. power industry.

Cat-Ox systems are provided in multiple, parallel independent trains, so that part of the power plant can continue to operate even if one train should fail. The Cat-Ox reheat system can be bypassed in the event of complete failure. If, however, part or all of the integrated Cat-Ox system should fail, it is necessary to reduce the load on the boiler or shut it down. Consequently, the converter in each train is further divided into parallel modular sections that can be dampered off to permit catalyst cleaning if this should be necessary between scheduled shutdowns.

It appears possible to cope with the actual or potential technical problems of the Cat-Ox system should components of the existing system not prove adequate. However, the alternative solutions will tend to increase both capital costs and those operating costs other than maintenance.

The by-product produced by the Cat-Ox system, 78 percent sulfuric acid contaminated with a small amount (less than 0.1 percent) of fly ash, is not a generally favorable one from an economic standpoint. Unless it can be transported by barge, its movement to markets is severely limited by the cost of shipment. This is also true, though in lesser degree, of the more concentrated acids, but the Cat-Ox product acid is additionally limited in applications by its lower concentration and its impurities. The manufacture of phosphate fertilizers is virtually the only large and growing market that might absorb a large output of the Cat-Ox acid, and the availability of even this market is highly dependent upon specific local conditions.

To permit estimates of capital and annual costs of the Cat-Ox and Wellman-Lord SO_2 Recovery systems, two models of hypothetical power plants were created. (Details are presented in Appendix A.) The first (Model A) is an existing 500-megawatt plant located in central Pennsylvania (Altoona). The second (Model B) is a new 1000-megawatt plant located on a navigable river in the Midwest (Cairo, Illinois). Cost estimates were made for a Cat-Ox reheat system to be applied to the Model A plant (Table I) and for an integrated Cat-Ox plant to be applied to the Model B Plant (Table II).

Table I

SUMMARY OF ESTIMATED COSTS
CAT-OX REHEAT SYSTEM FOR POWER PLANT
MODEL A

Item	Cost or Credit
Capital Investment	$16,200,000 $32.40/kw
Gross Costs before Credits	$4,853,950/yr 1.39 mills/kwh 13.9¢/million Btu $3.70/ton of coal
Credits for Product Acid[1] A. Acid price $11.00/ton B. Acid price $ 6.00/ton	 $1,181,000/yr $ 644,300/yr
Net Costs after Credits A. Acid price $11.00/ton	 $3,672,750/yr 1.05 mills/kwh 10.5¢/million Btu $2.80/ton of coal
B. Acid price $6.00/ton	$4,209,650 1.20 mills/kwh 12.0¢/million Btu $3.20/ton of coal

[1] Prices on 100-percent acid basis.

Table II

SUMMARY OF ESTIMATED COSTS
INTEGRATED CAT-OX SYSTEM FOR POWER PLANT MODEL B

Item	Cost or Credit
Gross Capital Investment	$50,800,000
Capital Credit	$16,896,000
Incremental Capital Investment	$33,904,000
	$33.90/kw
Gross Costs before Credits	$ 7,610,940/yr 1.09 mills/kwh 11.3¢/million Btu $3.02/ton of coal
Credit for Product Acid[1] Acid price $10.00/ton	$2,084,600/yr
Net Costs after Credits	$5,526,340/yr 0.789 mill/kwh 8.22¢/million Btu $2.19/ton of coal

[1] Price on 100-percent acid basis.

Estimates of annual cost were based on a plant life of 20 years and on an annual operating time of 7000 hours (load factor 80 percent). The estimated prices for by-product sulfuric acid are based on an assumed Gulf Coast price for sulfur of $30/long ton, which is probably as high as can be anticipated in the period up to 1975.

For the Cat-Ox reheat system used in the Model A plant, fixed charges account for 48 percent of the gross total annual cost before allowance for by-product credits. The next largest item of cost (18 percent) is that for the No. 2 fuel oil used in reheating the flue gas. Two acid prices are presented in estimating the by-product credit for the acid produced; the lower of the two is applicable in the event that it should be necessary to displace existing captive acid producers in the Pittsburgh area, where the acid from the Altoona plant would have to be sold.

For the integrated Cat-Ox system used in the Model B plant, a capital credit is due because a large part of the gross capital investment is for items that would be part of the conventional power plant without sulfur dioxide emission controls. Fixed charges account for 65 percent of the gross total annual cost before allowance for by-product credits. Maintenance (13 percent) and electrical power (10 percent) are the next largest cost items.

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IV PROCEDURES

A. Formulation of Models

The formulation of the models of the hypothetical power plants was based first on conditions specified by DPCE-NAPCA. Part of the remaining model conditions were derived from the NAPCA specifications so as to be consistent with the latter. Other conditions had to be selected on a relatively arbitrary basis, and where this was necessary, guidance was obtained from the literature or from discussions with informed individuals. In other cases, conditions had to be selected on completely arbitrary bases, and the author used his own experience and best judgment. The complete models are presented in Appendix A.

The basic premise employed throughout was that all estimates of the costs of sulfur oxides control should be made in relation to a base level represented by the equivalent conventional plant without sulfur oxides emission controls. Emission control was to be charged with whatever costs were incurred above those of the conventional, or base-level, plant. Similarly, it was to be credited with any savings from the cost of the base-level plant, or with any income derived from sale of by-products.

If the sulfur oxide control system included components common to the base-level plant, it was to be charged only with the increment in the costs of these components over those of the corresponding items in the base-level plant (or credited with the difference if the items in the conventional plant were more expensive).

The sulfur oxide control system was to be charged with the cost of power required to move the flue gas through any parts of the system specific to emission control. It was also to be charged with a proportionate share of the capital cost of the fan and motor if the latter also supplied draft for the rest of the boiler system.

The model formulations were circulated to DPCE-NAPCA and to Monsanto Company and Wellman-Lord, Inc. for review before adoption.

B. Cost Factors

The factors used in making cost estimates, and the bases on which they were adopted, are presented in Appendix B. Salaries, payroll benefits, and overhead were taken to be the same as those used in the previous study of the nonferrous smelting industry.[8]

Estimates of the prices that might be obtained for sulfur by-products were made by Stanford Research Institute, and are presented in Appendix D. The choices of the specific sites for the hypothetical plants (within the general areas specified by DPCE-NAPCA) were made to facilitate estimation of definite prices.

C. Preparation of Technical Data and Cost Estimates

The Monsanto Company prepared the technical designs for the model control systems, based on the conditions formulated by SRI, and estimated the capital investments and the utilities and maintenance requirements. The author reviewed these estimates and accepted most of them. Since the design data and specific component configurations were in most cases not revealed by Monsanto, it was generally impractical to make a critical, independent analysis of the capital cost estimates. The cost breakdowns for individual components of the systems were supplied by Monsanto, however, and were surveyed for general reasonableness and consistency with the model specifications.

Some of the original estimates of system requirements or costs made by Monsanto were not accepted, and the author supplied his own estimates. Most of the differences were resolved with Monsanto; the operating labor requirement was the only substantial exception.

V PROCESS DESCRIPTION

The Monsanto Cat-Ox process is essentially a variation on the contact process for sulfuric acid manufacture, which is extensively treated in the literature.[4] The process was developed primarily for removal of sulfur dioxide from the flue gases of coal- or oil-fired power plants,[14,15] in which the sulfur dioxide concentration usually falls within the range from 0.1 to 0.4 percent by volume. It differs from the conventional contact process in three principal respects:

1. Because of the diluteness of the sulfur dioxide, the plant is not autothermal; that is, the heat released by the oxidation of the sulfur dioxide is insufficient to preheat the feed gas to the reaction temperature (generally above 800°F). Hence, if the gas is not already near that temperature, auxiliary heat must be employed.

2. The system operates on wet gas. The feed gas is not dried before it enters the converter (catalytic reactor). Hence, the amount of sulfuric acid mist formed is relatively much greater than that formed in the conventional contact process, and a collector must be employed to recover the mist from the tail gas.

3. The heat in the exit gas from the converter is used to a greater or lesser degree to concentrate the sulfuric acid formed in the final absorption step. The concentration of the acid produced depends upon the temperature to which the flue gas stream is reduced during contact with the acid in the absorber.[14]

All of the component basic concepts of the Cat-Ox system have been demonstrated previously under some circumstances. Catalytic oxidation of sulfur dioxide at low concentrations has been demonstrated previously in the laboratory[11] and in pilot plant studies that were preliminary parts of the development of the Cat-Ox system.[15] A similar process (SNPA-Topsoe) has been in large-scale commercial operation in Lacq, France for several years on the incinerated tail gases from a Claus sulfur plant;[5,6,7] there the concentration of sulfur dioxide is of the order of 1.0 percent.

Some relatively conventional contact plants for manufacture of sulfuric acid from hydrogen sulfide have been built for use on wet gases.[4] The hot gas from the combustion of hydrogen sulfide (partly cooled in a waste heat boiler and diluted with air to give a sulfur dioxide concentration of about 7.5 percent) enters the converter without being dried. The hot exit gas from the converter is not cooled, as is customary in contact plants, but goes directly to an absorber. Because of the relatively high water content of the gas stream, it is not feasible to produce acid stronger than 93 to 94 percent. Also, relatively large quantities of sulfuric acid mist are formed that must be removed from the tail gas with some type of mist collector. The principal advantage sought in this type of plant is a reduction in capital cost.

In this report, two variations of the Cat-Ox system are described:

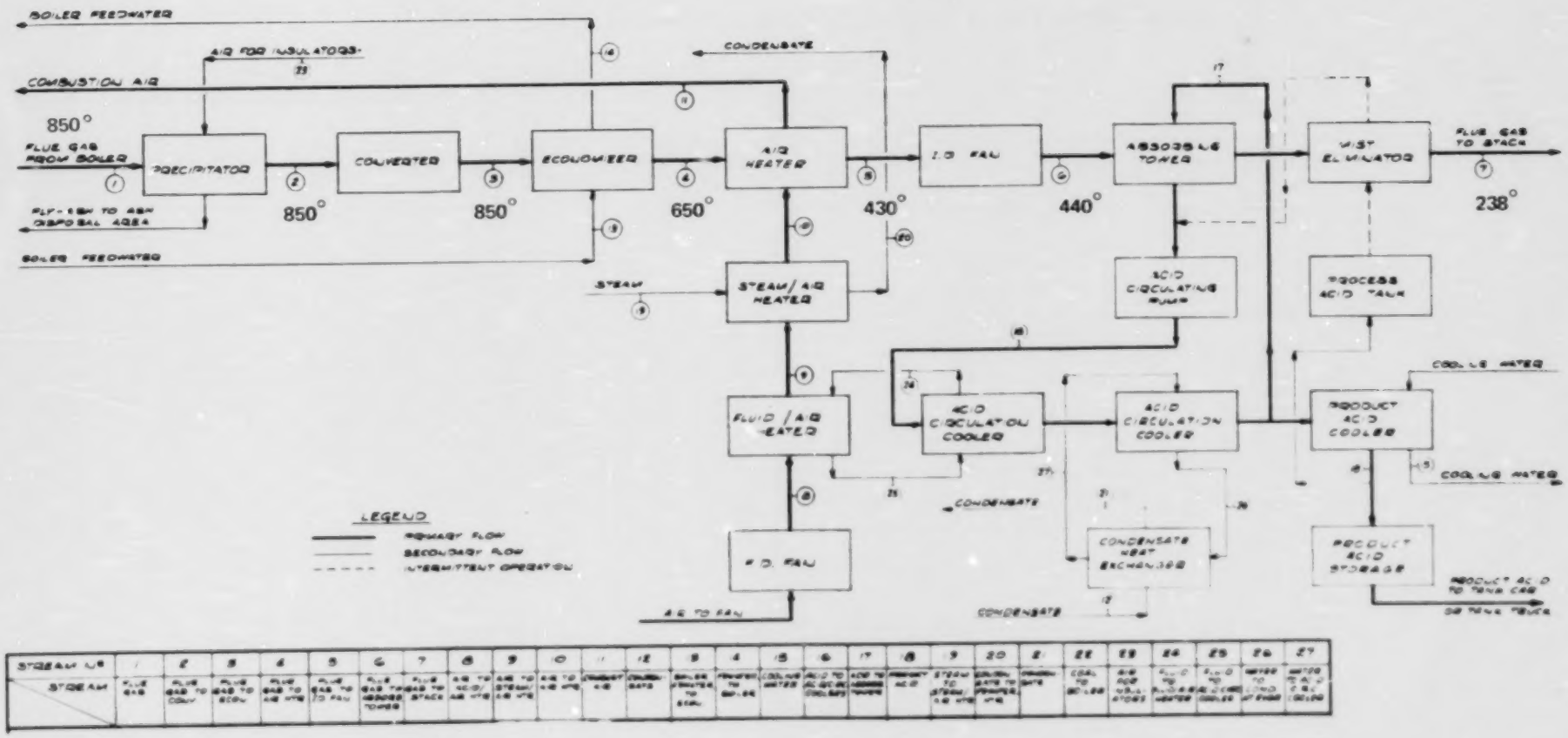
1. An integrated system to be constructed as part of a new power plant (Fig. 1).

2. A system to be added to an existing power plant (Fig. 2).

A. Integrated Cat-Ox System for Power Plants

The integrated Cat-Ox system, which is conceptually applied to the hypothetical power plant Model B, cannot be applied to an existing power plant without an extended shutdown and major alterations of the plant. Consequently, it will generally be feasible to install it only where it can be incorporated as an integral part of a new plant.

The system as applied in the demonstration plant at the Portland station of the Metropolitan Edison Company is described in a paper by Stites et al.[12] The arrangement as proposed for use in a large power plant typified by Model B differs in some details from that employed in the demonstration plant. Parts of the system, as shown in Fig. 1, are integral components of the power plant and would be in use even if the Cat-Ox system were not employed. These are the electrostatic precipitator, the fly ash handling system, the economizer, the air heater,



Courtesy of Monsanto Enviro-Chem Systems, Inc.

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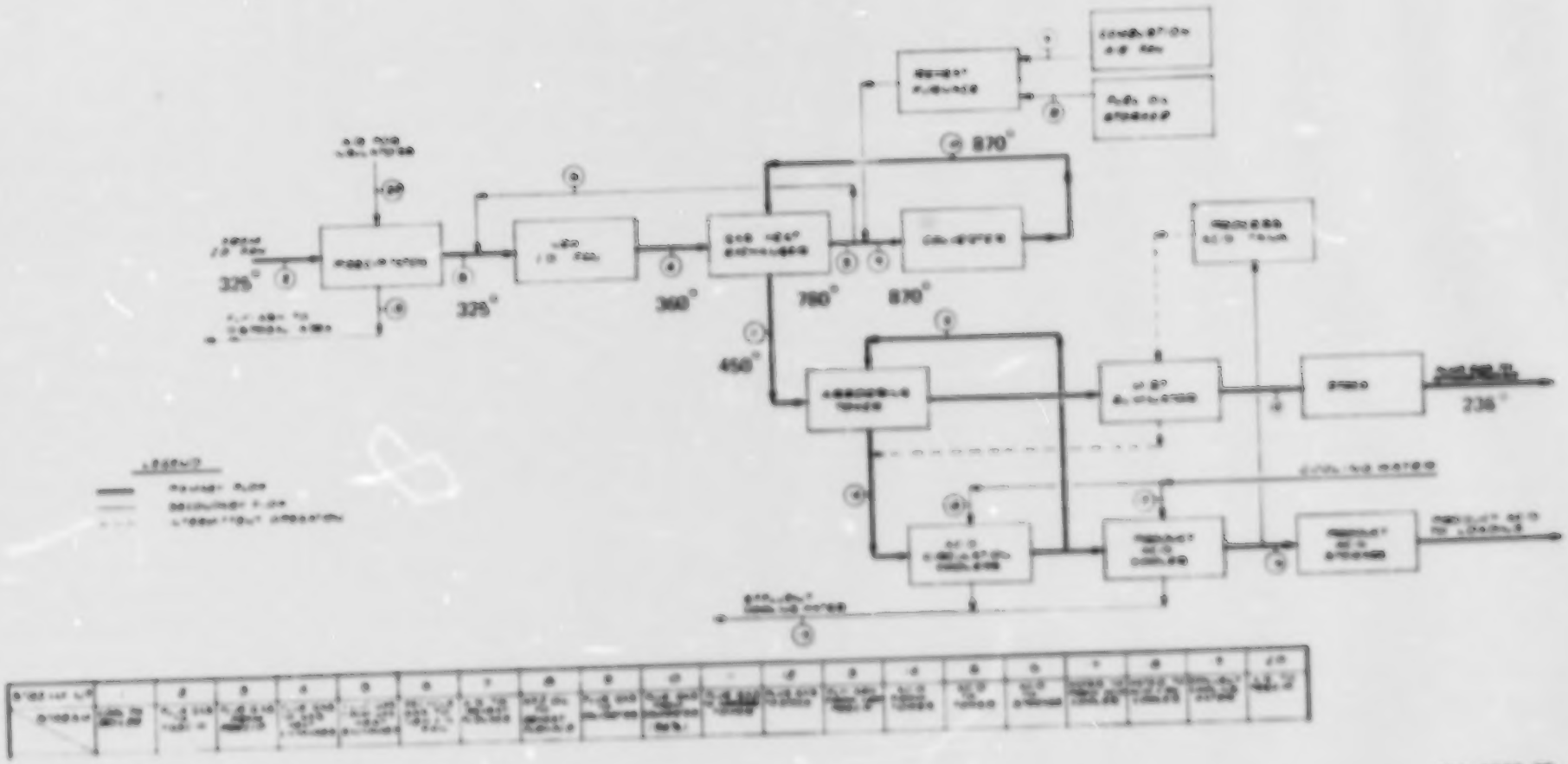
FIGURE 1 INTEGRATED CAT-OX SYSTEM FOR POWER PLANT MODEL B



Courtesy of Monsanto Enviro-Chem Systems, Inc.

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FIGURE 2 CAT-OX REHEAT SYSTEM FOR POWER PLANT MODEL A

the steam/air heater, the induced draft fan, and the forced draft fan. Some of these units are, however, more and some less costly than the corresponding ones in the conventional power plant. The electrostatic precipitator handles the flue gas at a higher temperature than in the conventional installation, and hence must be correspondingly larger. (Fly ash precipitation is reportedly easier at the higher temperature, however, so that the increase in the required precipitator size may not be directly proportional to the increase in gas volume.) The efficiency of the precipitator must also be higher than is currently being provided in any commercial fly ash precipitators.

The air heater recovers less of the heat from the flue gas than does that in a conventional plant, and hence is smaller than the latter. The induced draft fan must be sized and powered to overcome the gas flow resistance of the converter, absorbing tower, and mist eliminator as well as that of the elements of the conventional system.

The air heater proposed by Monsanto for the Model B plant is a tubular recuperative type. Alternatively, a regenerative air heater of the Ljungstrom type might be used. The Ljungstrom air heater, which is widely used in conventional power plants, is less costly than the tubular type. However, there is leakage of air into the flue gas. Hence, with the Ljungstrom unit the system downstream of the air heater would have to be designed to handle a slightly greater gas flow.

The outlet flue gas temperature at the air heater, and hence the amount of heat recoverable in the latter, is determined by the necessity for remaining above the acid dew point to avoid corrosion. In addition, it is necessary to avoid "cold-end" corrosion in the air heater itself. If air at ambient temperature is admitted directly into the air heater, the metal wall temperature may be low enough in localized areas for acid condensation to take place on the flue-gas side. To avoid this, the combustion air is heated first in the fluid/air heater, and second in the steam/air heater. Such tempering of combustion air with a steam/air heater is commonly practiced in conventional power plants using high-sulfur fuels, also in order to avoid air heater corrosion.

The fluid-air heater is employed to recover additional heat from the circulated acid, which is cooled after each pass through the absorbing tower. Heat is transferred from the acid circulation cooler to the fluid/air heater through an intermediate heat transfer fluid. Direct transfer of heat from the acid to the air in the fluid/air heater is avoided because a leak in the tubing could introduce acid into the combustion air system.

Additional heat is recovered from the acid by heating of condensate as feed water for the boiler. The possibility of acid leaking into the boiler feed water is avoided by circulation of an intermediate heat transfer fluid between the acid circulation cooler and the condensate heat exchanger.

The heat recovered from the acid by the preheating of the boiler feed water and combustion air compensates for the reduced heat recovery in the air heater. The overall heat recovery in the power plant cycle is essentially the same as in the corresponding conventional power plant, although part of the heat is recovered at lower temperature levels at a corresponding increase in capital expenditure for heat transfer equipment. The detailed breakdown of the low-level heat recovery is considered confidential by Monsanto, and stream temperatures in this part of the system are therefore not indicated in Fig. 1. The minimum outlet gas temperature at the absorbing tower is set by the concentration of the acid produced (78 percent).

In a conventional power plant, the preheating of the boiler feed water is accomplished with steam extracted from the turbines driving the generators. In the Cat-Ox system, where the boiler feed water is preheated in the acid cooler, the equivalent amount of steam can be used directly to drive the turbines, providing a credit for additional power generation. However, additional cooling water must be supplied to the condensers to condense the additional amount of steam exhausted from the turbines. The cost of the extra cooling water is taken as a debit against the Cat-Ox system.

The absorption of the sulfur trioxide in the cooled, recirculated acid is accomplished in a packed tower. The mist of sulfuric acid formed during the cooling of the gas, plus any entrained droplets of the circulating acid carried out of the tower, are recovered in a fiber-bed mist eliminator. Fine particles of fly ash that have escaped collection by the electrostatic precipitator or the catalyst bed in the converter are collected in part in the absorber, and remain in the product acid. The particles not removed from the gas in the absorber are collected with very high efficiency by the mist eliminator. Other fly ash particles present in the circulating acid may be carried in entrained droplets to the surfaces of the mist eliminator. The gradual buildup of solid particles in the eliminator produces an increase in the gas pressure drop, which must be counteracted by periodic washing of the fiber beds. The maximum permissible solids buildup is determined by the allowable gas pressure drop, which is in turn set by the fan and motor capacity provided for in the system design. In the design used, the pressure drop across the mist eliminator ranges from approximately 13 inches of water in the clean condition to 20 inches of water with the maximum permissible solids buildup.

The original mist eliminator used in the Portland demonstration plant was a Brink mist eliminator.[2,12] It was necessary to shut down the unit in order to clean the mist eliminator elements when they had become plugged with fly ash particles. However, Monsanto now has under development a new type of fiber-bed eliminator that is designated as the Cat-Ox mist eliminator.[9,10] Monsanto reports that this unit can be washed while on stream, eliminating the need for shutdowns except at scheduled maintenance periods. As provided for in the systems shown in Figs. 1 and 2, the fiber beds are washed with a portion of the product acid that is continuously filtered and accumulated in the process acid tank. Monsanto states that the fiber beds may be washed with water instead of acid, but has presented no estimates of the probable frequency of washing required under average operating conditions. If the frequency of washing is not too great and the quantity of flushing fluid required is moderate, it should be possible to use water without excessively diluting the product acid.

Monsanto considered the design and construction of the Cat-Ox mist eliminator to be particularly sensitive information falling outside the provisions of the confidentiality agreement, and did not reveal the information for evaluation in this study.

The Cat-Ox system as proposed for the Model B plant consists of four parallel trains, each of which can be operated independently. Modular construction is employed; each converter is divided into three sections which can be closed off independently by dampers. Thus, one-twelfth of the total converter flow area can be cut out of the system if it should be necessary to clean the catalyst between the regular plant shutdown periods.

The gradual buildup of fly ash in the catalyst bed results in an increase in the gas pressure drop, although there is reportedly little effect on the conversion efficiency of the catalyst. The system therefore includes equipment to withdraw, convey, clean, and return the catalyst to the converter. Monsanto estimates the total downtime for a converter unit as two or three days, including time for cool-off, cleaning, and heat-up of the catalyst. The loss of catalyst due to attrition and breakage during screening and mechanical handling is estimated by Monsanto to be about 2.5 percent by volume (and weight) for each cleaning. Monsanto also anticipates that it will be necessary to clean the catalyst beds after each three months of operation, or a total of four times per year. The corresponding loss of catalyst is about 10 percent per year. The vanadium pentoxide catalyst is a special type designed specifically for use in the Cat-Ox system.

In the converter, a single pass through a catalyst bed is employed to attain 90-percent conversion of the sulfur dioxide to the trioxide with the boiler furnace operating at full load. Because power consumption represents such a large item in the operating cost, the catalyst bed depth and configuration are designed for low gas pressure drop (1 inch of water when clean). As with the mist eliminator, the maximum permissible solids buildup in the catalyst bed is set by the fan and motor capacity, and the Monsanto design provides for a normal maximum pressure drop

through the dirty bed of 4 inches of water. The required precipitator efficiency and the permissible dust concentration in the gas stream entering the converter are considered by Monsanto to be confidential information. It can be stated only that the precipitator efficiency is in excess of 99.6 percent.[9]

The principal incremental gas pressure drop chargeable against the Cat-Ox system is that through the converter, absorbing tower, and mist eliminator. The pressure drops across the economizer, air heater, and electrostatic precipitator would be incurred in the conventional system. The pressure drop across the absorbing tower is approximately 5 inches of water, and there are some losses in the additional ductwork of the Cat-Ox system. The total incremental pressure drop is taken as 22 inches of water for the clean system and 32 inches of water with the designed maximum allowable solids buildup.

The conversion of sulfur dioxide to the trioxide in the contact process depends on both the temperature and the contact time in the catalyst bed.[4] In the integrated Cat-Ox process, variations in boiler load will affect both the gas temperature and the contact time. As the boiler load is reduced, the flue gas temperature tends to drop[12] but the contact time increases, so that the two effects tend within limits to counteract one another. The Monsanto design is intended to provide a high level of conversion over a reasonable range of variation in the boiler load. However, it is anticipated that a reduction of the boiler load to 50 percent of the rated capacitity may reduce the flue gas temperature to about 750^{o} and reduce the conversion of the sulfur dioxide to about 80 percent.[12] In such a case, the quantity of sulfur dioxide emitted would remain the same as at full load, because the reduction in the quantity of flue gas would compensate for the reduced conversion efficiency.

B. Cat-Ox System for Application to Existing Power Plant

The Cat-Ox system proposed for application to the Model A power plant is additive instead of integral. It could be bypassed entirely without affecting the operation of the power plant except during a rela-

tively short period in which the final tie-ins are made; this period could coincide with the annual shutdown of the power plant for overhaul.

The additive Cat-Ox system (Fig. 2), termed by Monsanto a "reheat system," takes the gas from the discharge of the existing induced draft fan. A new precipitator is added to reduce the quantity of dust to a tolerable level, which is the same as in the integrated Cat-Ox system. A new induced draft fan is added to move the gas through the Cat-Ox system.

The relatively cool (325°F) flue gas must be heated to a temperature high enough to permit catalytic oxidation of the sulfur dioxide. Part of this preheating is accomplished by exchange of heat from the converter exit gas in the gas heat exchanger. The rest of the preheating is accomplished by direct injection of hot combustion gases from the reheat furnace, which is fired with No. 2 fuel oil of low ash content. The hot gas leaving the converter passes through the gas heat exchanger where it heats the incoming cool feed gas and is itself cooled to 450°F before it enters the absorbing tower. From this point on, the arrangement and operation are essentially the same as in the integrated Cat-Ox system. The temperature of the gas leaving the heat exchanger, 450°F, is set as in the integrated system by the necessity of staying above the sulfuric acid dew point.

In the gas heat exchanger, as in the air heater of the integrated system, it is necessary to avoid cold-end corrosion. The temperature of the incoming cool flue gas must therefore be raised somewhat, and this is accomplished by the recycle of a portion of the hot exit gas to the inlet of the exchanger. The sizes of the new induced draft fan and the gas heat exchanger must be increased accordingly to handle the greater total flow of gas.

The acid circulated to the absorbing tower is cooled directly with cooling water, and there is no recovery of the low-level heat. The product acid is of 78 percent concentration, as in the integrated system.

The gas heat exchanger is of the Ljungstrom regenerative type instead of the recuperative type employed as the air heater in the inte-

grated system. In the regenerative heater, seal leakage permits some of the cool incoming gas to bypass the heater and converter and infiltrate the hot gas stream going to the absorbing tower. The sulfur dioxide in the leakage gas is not oxidized and recovered. Consequently, the fractional conversion of sulfur dioxide in the gas that does pass through the converter must be increased to compensate. Monsanto estimates that the leakage will be 5 percent of the input gas flow, and that 94 percent conversion of the sulfur dioxide can be maintained. Overall recovery of sulfur dioxide as sulfuric acid will be 89 percent.

In the reheat system as contemplated for general application to power plants,[10] Monsanto proposes to design for 90 percent conversion, so that with a leakage of 5 percent of the gas through the heat exchanger the overall removal of sulfur dioxide will be 85 percent. However, for the Model A plant an overall removal of 90 percent of the sulfur dioxide was specified. In the Cat-Ox converter design to be used for power plant applications, 94 percent conversion is a practical upper limit; hence, the overall removal of sulfur dioxide falls slightly short of the specified level, although not very significantly so. The converter could be designed to achieve a greater conversion, but presumably it would require a greater quantity of catalyst and have a higher gas pressure drop. A preferable alternative approach might be to use a tubular heat exchanger, which would be more costly but would avoid the gas leakage problem associated with the Ljungstrom exchanger.

In the Cat-Ox plant proposed by Monsanto for application to the Model A power plant, provisions are made to divert the entire flow of flue gas directly to the stack through a dampering system should it be necessary to take the recovery system off the line for unscheduled maintenance or catalyst cleaning. Hence, no provisions are made in the converter for taking modular sections out of service for catalyst cleaning while the gas flow is continued through the remaining sections, as can be done in the integrated system. Scheduled catalyst cleaning will take place at three-month intervals, and periodic flushing of the mist eliminator can be carried out, as in the integrated system.

Two complete and independent parallel trains are used, so that one-half of the normal full-load gas flow can be accommodated if one train is out of service.

The equipment and its arrangement for reheating the flue gas and subsequently recovering heat from it[10] were chosen by Monsanto to achieve desired economies in capital investment and operating costs. The use of No. 2 fuel oil for reheating was specified by SRI for the Model A plant. Natural gas might be substituted where it is available. Coal or residual oil could be used in a reheat system, but the equipment and its arrangement would necessarily be different than are portrayed in Fig. 2 and Reference 10.

The entire capital and operating costs of the additive, or reheat, Cat-Ox system are, of course, chargeable to sulfur dioxide recovery. The total gas pressure drop through the system ranges from 31 inches of water in the clean condition to 41 inches of water with the designed maximum permissible dust buildup in the converter and mist eliminator.

In the reheat Cat-Ox system, the temperature of the flue gas entering the converter is independent of the boiler load and can be held at a constant level even while the boiler load and the flow of flue gas are varying. Hence, the conversion can be maintained at a high level at all times. A reduction in flue gas flow should permit an increase in conversion because of the increased contact time in the catalyst bed.

VI PROCESS DATA AND COST ESTIMATES

A. Material Flows

The estimated quantities of coal burned in the two hypothetical power plants -- and hence the calculated quantities of flue gas, sulfur compounds, and fly ash to be handled -- were fixed by the assumed heat rates (see Appendix A). The heat rates assumed for the Model A and Model B power plants were 10,000 and 9,600 Btu/kwh, respectively. These values are probably between 5 and 10 percent higher than those that might reasonably be obtained at comparable actual plants. Although the assumed heat rates are perhaps excessively conservative, they are well within the probable precision of the other estimates made for the study.

Although the calculated volume of flue gas generated was increased by use of the high values of heat rate, no allowance was made for any increase in gas volume resulting from air infiltration into the flue gas handling system, which occurs in actual practice.

The estimated rates and conditions of gas flow through the Model A and Model B Cat-Ox systems are presented in Tables III and IV, respectively. The flow rates of the main flue gas stream are given both at standard conditions (70°F, 1 atm) and at the actual conditions existing at the indicated points in the system. Allowances were made for the small quantities of air that bleed into the system after being used to purge the insulator compartments of the electrostatic precipitators. In the Cat-Ox reheat system (Model A), the flue gas also receives a small additional quantity of gas from the direct-fired reheater furnace.

Some water vapor is removed from the flue gas in the absorbing tower, where it reacts with sulfur trioxide to form sulfuric acid and also dilutes the product.

The estimated distributions of the contaminants (sulfur oxides and fly ash) in the control systems are presented in Tables V and VI. In the

Table III

MODEL A
GAS FLOWS IN CAT-OX REHEAT SYSTEM

Gas Flows Entering System	
Flue gas entering precipitator	1,564,360 CFM (325°F, 1 atm)
	1,056,200 SCFM (70°F, 1 atm)
Purge air to precipitator insulator compartments	4,510 SCFM (70°F, 1 atm)
Flue gas from reheater furnace	35,280 SCFM (70°F, 1 atm)
Total	1,095,990 SCFM (70°F, 1 atm)
Gas Flows Leaving System	
Water vapor in product acid	5,110 SCFM (70°F, 1 atm)
Tail gas to stack	1,432,550 CFM (236°F, 1 atm)
	1,090,880 SCFM (70°F, 1 atm)
Total	1,095,990 SCFM (70°F, 1 atm)

Table IV

MODEL B
GAS FLOWS IN INTEGRATED CAT-OX SYSTEM

Gas Flows Entering System	
Flue gas entering precipitator	5,011,840 CFM (850°F, 1 atm)
	2,027,690 SCFM (70°F, 1 atm)
Purge air to precipitator insulator compartments	15,500 SCFM (70°F, 1 atm)
Total	2,043,190 SCFM (70°F, 1 atm)
Gas Flows Leaving System	
Water vapor in product acid	9,930 SCFM (70°F, 1 atm)
Tail gas to stack	2,677,760 CFM (238°F, 1 atm)
	2,033,260 SCFM (70°F, 1 atm)
Total	2,043,190 SCFM (70°F, 1 atm)

Table V

MODEL A
CONTAMINANT REMOVAL IN CAT-OX REHEAT SYSTEM

Contaminants Entering in Flue Gas	
Fly ash	1,405 lb/hr
SO_2	22,200 lb/hr
SO_3	391 lb/hr
Contaminants Emitted in Tail Gas	
Fly ash	Negligible
H_2SO_4 vapor and mist	142 lb/hr
SO_2	2,380 lb/hr
By-Product Recovered	
78-percent H_2SO_4	39,330 lb/hr
Equivalent 100-percent H_2SO_4	30,680 lb/hr

Table VI

MODEL B
CONTAMINANT REMOVAL IN INTEGRATED CAT-OX SYSTEM

Contaminants Entering in Flue Gas	
Fly ash	53,960 lb/hr
SO_2	42,620 lb/hr
SO_3	751 lb/hr
Contaminants Emitted in Tail Gas	
Fly ash	Negligible
H_2SO_4 vapor and mist	252 lb/hr
SO_2	4,160 lb/hr
By-Product Recovered	
78-percent H_2SO_4	76,360 lb/hr
Equivalent 100-percent H_2SO_4	59,560 lb/hr

estimates for the Model A system (Table V), the very small quantity of sulfur present in the No. 2 fuel oil used for reheating has been ignored. For the amount of sulfuric acid emitted in the tail gases from the systems, Monsanto did not specify the distribution between vapor and mist. However, most of the escaping acid is in the vapor state; probably no more than 10 percent of it is mist.

The quantities of fly ash emitted in the tail gases are characterized as negligible. Estimates of the actual quantities were not provided by Monsanto; however, the author believes that the emissions of fly ash will not exceed three to five lb/hr.

B. Capital Cost Estimates

The estimates of capital investment presented in Tables VII, VIII, and IX were prepared by Monsanto with only two minor exceptions. The author modified the original credit for the electrostatic precipitators of the Model B plant (Table VIII) for better conformity with the specifications set for the models. The credit represents the cost of the precipitator of 99.5 percent efficiency, operating at 325°F, that would be used on the power plant if no sulfur oxide emission controls were employed (see Appendix A).

The author also provided the credit for the stack for the Model B plant, which represents the difference between the cost of an 800-foot stack used with no sulfur oxide emission controls, and that of a 500-foot stack used when emission controls are applied (see Appendix C). The same stack credit was assigned arbitrarily to both the Cat-Ox and Wellman-Lord control systems for the Model B plant. The precision of the estimate is low, and the credit is provided mostly for purposes of illustration. As is discussed in Appendix C, some reduction of stack height in this model case appears justified. It appears to the author, as to some others,[1] that the problems of securing dispersion of scrubbed stack gases have been overstressed.

Table VII

POWER PLANT MODEL A
SUMMARY OF CAPITAL AND OPERATING COSTS FOR CAT-OX REHEAT SYSTEM

Item	Quantity	Cost Basis	Cost
Capital Investment[1]		$	16,200,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	2,349,000
B. Direct Operating Costs			
1. Operating labor	9333 hrs/yr	$3.75/hr	35,000
2. Supervision	1750 hrs/yr	$4.75/hr	8,310
3. Payroll benefits		25% of labor and supervision	10,830
4. Maintenance		3% of capital investment	486,000
5. Electricity	15,100 kw	0.55¢/kwh	581,350
6. Fuel oil	1200 gal/hr	$4.25/barrel	850,000
7. Cooling water	32,000 gpm	$0.02/1000 gal	268,800
C. Indirect Costs			
1. Overhead		50% of labor, supervision, and maintenance	264,660
Total Annual Cost (Before credits)		$	4,853,950
		mills/kwh	1.39
		¢/million Btu	13.9
		$/ton coal	3.70
Credit for Acid (100%)	107,380 tons/yr	$11.00 ton 100% acid	1,181,200
		$ 6.00 ton 100% acid	644,300
Net Annual Cost (After credits)			
A. Acid Price $11.00/ton		$	3,672,750
		mills/kwh	1.05
		¢/million Btu	10.5
		$/ton coal	2.80
B. Acid Price $6.00/ton		$	4,209,650
		mills/kwh	1.20
		¢/million Btu	12.0
		$/ton coal	3.20

[1] Does not include land, spares, interest on investment during construction, start-up expense, working capital, or royalty on process.

Table VIII

POWER PLANT MODEL B
CAPITAL INVESTMENT FOR INTEGRATED CAT-OX SYSTEM

Total Capital Investment (Before Credits)	$50,800,000
Credits	
Induced Draft Fans and Drives	963,000
Forced Draft Fans and Drives	738,000
Precipitators and Supports	3,100,000
Air Heaters	3,195,000
Economizers	716,000
Fly Ash System	253,000
Flues, Ducts and Supports	3,007,000
Steam Coil Air Heaters	67,000
Electrical Switchgear and Wiring	598,000
Stack	850,000
General Field Costs	2,100,000
Subtotal	$15,587,000
Engineering and Engineering Overhead	530,000
Contingencies @ 5%	779,000
Total Credits	$16,896,000
Incremental Cost of Cat-Ox System	$33,904,000

Table IX

POWER PLANT MODEL B

SUMMARY OF CAPITAL AND OPERATING COSTS FOR INTEGRATED CAT-OX SYSTEM

Item	Quantity	Cost Basis	Cost
Capital Investment[1]		$	33,904,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	4,916,100
B. Direct Operating Costs			
1. Operating labor	18,666 hrs/yr	$3.75/hr	70,000
2. Supervision	3,500 hrs/yr	$4.75/hr	16,630
3. Payroll benefits		25% of labor and supervision	21,660
4. Maintenance		3% of capital investment	1,017,100
5. Electricity	20,700 kw	0.55¢/kwh	796,950
6. Cooling water	466 gpm	$0.02/1000 gal	3,910
7. Condenser cooling water	25,800 gpm	$0.02/1000 gal	216,720
C. Indirect Costs			
1. Overhead		50% of labor, supervision and maintenance	551,870
Total Annual Cost (Before credits)		$	7,610,940
		mills/kwh	1.09
		¢/million Btu	11.3
		$/ton coal	3.02
Credit for Acid (100%)	208,460 tons/yr	$10.00/ton 100% acid	2,084,600
Net Annual Cost (After credits)		$	5,526,340
		mills/kwh	0.789
		¢/million Btu	8.22
		$/ton coal	2.19

[1] Does not include land, spares, interest on investment during construction, start-up expense, working capital, or royalty on process.

C. Operating Cost Estimates

The estimates of fuel, water, and power requirements for the two Cat-Ox control systems (Tables VII and IX) were developed by Monsanto, and the corresponding operating costs were calculated by application of the appropriate cost factors.

The estimate of the annual maintenance cost as three percent of the Cat-Ox plant capital investment was provided by Monsanto; it includes the cost of catalyst makeup. However, this estimate is predicated upon maintenance of specified standards of fly ash collector performance, and if the latter are not consistently attained, the costs for maintenance of the equipment may be substantially higher, as is discussed in Sections V and VII.

Monsanto proposes that the power plant operators should also operate the Cat-Ox system, and that the only extra labor required should be that for handling product acid on the day shift. The author believes that at least one full-time operator will be required on each shift for the Model A Cat-Ox plant, and that at least two will be required for the Model B Cat-Ox plant. These operating labor allowances have therefore been included in the cost estimates presented in Tables VII and IX. In addition, one day laborer is provided for the Model A plant and two for the Model B plant. The author considers that this operating labor allowance is a minimum, and that in view of the critical nature of the fly ash collector performance, the actual requirement might be twice as great.

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VII GENERAL DISCUSSION

A. Evaluation of the Cat-Ox System

There are no reasons to doubt the technical feasibility of the basic Cat-Ox process. As is noted above, the basic process (essentially, the contact process) is fully established. Technical problems remain with respect to specific aspects of application of the process, but these do not constitute essential limitations on the process itself. It is still uncertain whether some of these technical problems have as yet been resolved in a fully satisfactory manner, or whether the approaches taken are in all cases the most appropriate. However, none of these problems appears to be insoluble. The principal question concerns the possible additional costs that may be necessary to provide satisfactory solutions.

There are three principal technical problems (actual or potential) in the Cat-Ox process; these are listed below in their order of importance:

1. Dust collection
2. Acid mist formation and collection
3. Corrosion.

Of these three, corrosion is dealt with most readily. It presents its greatest threat in heat exchangers, where it can be avoided, but only at the expense of reduced heat recovery and of increased capital and operating costs. Dust collection is by far the greatest problem. Acid mist collection presents a problem primarily because of its interrelationship with the dust collection problem.

It is clear that a large part of the difficulties involved in the development of the Cat-Ox process -- probably the great majority of all the difficulties -- have been associated with fly ash. The permissible

quantity of dust in the gas entering the converter represents a balance between the cost of gas cleaning and the cost of maintenance -- catalyst cleaning and makeup, and cleaning of the mist eliminator -- plus costs or losses associated with contamination of the product acid. The fly ash precipitator must operate not only with very high efficiency but also with a very high degree of reliability. A serious precipitator failure could speedily cause the catalysts beds to be loaded with fly ash to the point where the system would be inoperable. The problem of reliability is particularly acute with the integrated Cat-Ox system, which cannot be bypassed. Multiple trains and modular construction are necessary to ensure that most of the available boiler capacity will remain available even if a portion of the Cat-Ox system should fail. Nevertheless, even a deterioration of precipitator performance without outright failure may seriously increase the problems of converter maintenance. It will therefore be necessary to attain high standards of maintenance and operation of the precipitators -- probably much higher than the standards normally reached in the U.S. power industry. Providing adequate monitoring of precipitator performance may prove demanding, since even a slight deterioration in efficiency that would not be readily apparent might double the amount of dust entering the converter.

The collection of the acid mist itself is not likely to present any serious problems, since fibrous mist eliminators of the Brink type have been amply demonstrated to give both efficiency and reliability in mist collection. However, the deposition of residual fine fly ash particles in the packed fiber bed is a serious problem. Dislodging and removing solid particles deposited in depth in such packed beds is difficult, and the history of past attempts to do so is not encouraging. Solids buildup in the Brink eliminator proved to be a problem in the demonstration plant,[12] and it appears that no satisfactory solution has been developed until very recently, if indeed it has been attained even now.

Although Monsanto representatives declined to reveal any design information on the new Cat-Ox mist eliminator, they did permit the writer to make a brief inspection of some operating records on the unit being tested at the Portland demonstration plant. The writer believes that he

has been able to deduce the basic approach being taken by Monsanto. If this deduction is correct, the method being used is neither novel nor original, but it does possess merit. However, the degree of success attainable with this type of approach is acutely dependent upon the quantity and particle size distribution of the dust as well as upon the detailed design of the eliminator. These same factors can also have major effects upon the capital cost of the eliminator unit. Since information of this type has not been made available for study, it is not possible to evaluate the actual status of this development. It appears that substantial progress has been made with the new eliminator, but the writer is not convinced that there has yet been an adequate demonstration of the ability to keep the eliminator free of solids buildup over long periods of continuous operation.

If the approach being taken by Monsanto should not in fact be sufficiently effective, there are other techniques available for approaching the problem of dealing with the combination of acid mist and residual fine solid particles. It should be possible to develop adequate alternative solutions, although some additional capital and operating costs may be incurred.

From an economic standpoint, the Cat-Ox system presents some evident problems. The capital investment appears relatively high even when it is considered that cost estimates for some potential alternative systems may be highly optimistic. However, a greater limitation may be the nature of the by-product. The relatively dilute (78 percent) sulfuric acid is a material with a value acutely sensitive to plant location. It has few uses capable of absorbing a large and growing volume; phosphate fertilizer manufacture is the most important if not the only one. The acid cannot be economically shipped for substantial distances except by water. (See below and Appendix D.) The presence of residual fly ash particles in the acid may be objectionable in some possible applications, but should not be a significant limitation on use of the acid in phosphate fertilizer manufacture.

The problem of finding markets for the sulfuric acid is shared by any other processes for recovering acid -- even concentrated acid --

from power plant flue gases, but with the weak acid it is proportionately more serious. Sites for possible installations of the system will have to be chosen carefully; if several large power plants in a given area do install Cat-Ox systems, they could easily flood any feasible acid markets.

On the basis of this study, the technical improvements in the Cat-Ox system that appear most needed are ones that might increase the efficiency and reliability of the fly ash collection system. Potential improvements of this nature would probably add further to the capital cost, but might well be justified by reduction in the maintenance costs. As an example, cyclone collectors might be installed downstream of the electrostatic precipitator. Stanford Research Institute and other investigators have noted that a large fraction (half or more) of the dust penetrating dry precipitators may consist of reentrained flocs that are of relatively large particle size even though composed of smaller discrete particles. Properly designed secondary cyclones might cut the normal dust load entering the converter by as much as 50 to 75 percent. The material collected by the cyclones would be that which would deposit predominantly in the catalyst bed. Very fine discrete particles, which would tend to pass through the catalyst bed, would be little affected. The secondary cyclones would be of little benefit in the event of precipitator failure.

B. By-Product Values

Over the period of about one year (1969), sulfur supplies have passed abruptly from shortage to surplus, and it is now difficult, if not impossible, to determine whether there is an established price for sulfur. The Gulf Coast price of sulfur produced by the Frasch process has long been the base level for world sulfur prices, but it is apparently not so at present, and it is uncertain whether it will be again. In the previous study of the nonferrous smelting industry[8] it was estimated that the average Gulf Coast price of sulfur over the period to 1975 might be $30/long ton. When the estimate was made, the Gulf Coast price was over $40/long ton; currently, sulfur is reported to be selling in some areas for less than $15/long ton. Although sulfur prices could rise again during the next five years, the figure of $30 now appears to be at the

optimistic end of the probable range of prices that may prevail to 1975 (see Appendix D). Nevertheless, it has been used as the basis for by-product price estimates in this study because there appears to be none that can be used with significantly greater confidence, and the results still have some usefulness if their limitations are understood.

Even the original estimate of $30/long ton f.o.b. the Gulf Coast was based on the assumption that no very substantial quantity of sulfur would be recovered at power plants before 1975. The amounts of sulfur and sulfur by-products potentially recoverable from the coal consumed by power plants (whether from the coal itself or from the flue gases) are so large that recovery of a substantial part of them could completely disrupt existing local sulfur markets. The estimates made of sulfur and sulfuric acid prices that might be obtained by the Model A and Model B power plants (see Appendix D) embody the tacit assumption that no other large power plants near the assumed locations will also be recovering sulfur by-products.

In summary, it appears that the credits assigned to sulfur by-products from the Model A and Model B plants are the highest that can be reasonably expected in the period up to 1975. The actual credits might be much lower, and the situation that may exist after 1975 is unpredictable on the basis of currently available information. Conservatively, the long-term assessment of the economics of recovery processes should probably be based on the gross costs without allowance for by-product credits.

It is sometimes suggested that recovery processes can be more economically applied to power plant flue gases if coals with extra high sulfur contents are used. Since the sulfur dioxide concentration in the flue gas will be increased, the unit cost of production of the sulfur by-product should indeed be reduced. However, if the by-products must enter markets that are already glutted, the additional quantity of material may merely exert an increased downward pressure on prices. The actual advantage, if any, of using the high sulfur coal would probably be related to lower purchase prices for the coal.

C. Variations of Bases of Cost Estimates

The choice of some of the economic factors assumed in making the cost estimates was to a considerable degree arbitrary, as it has been in similar studies made by other workers. There is no general agreement among different workers on what constitutes a realistic set of bases. The result is that estimates for different control systems have been made on a wide variety of bases -- frequently only vaguely defined -- so that the results cannot be directly compared. In this study, it has been assumed that valid comparison of different control systems is more important than absolute accuracy, and hence that it is more important that the bases used should be uniform than that they should be the "correct" ones -- whatever the latter might be. An effort has been made to describe fully all bases used, so that the results can be recalculated to any different bases that other persons may wish to use. Although the bases chosen for the models (Appendixes A and B) are believed to be generally reasonable, it must be noted that other possible assumptions regarding some factors might be equally reasonable but lead to markedly different cost estimates.

1. Load Factor

The assumed annual load factor of 80 percent (equivalent to 7,000 hours per year of operation at full load) was specified by DPCE-NAPCA. It appears to be fairly reasonable for large, new, or relatively new baseload plants such as those typified by Models A and B. It is less certain whether the 80 percent factor can be assumed to hold over the whole of the assumed life (20 years) of the emission control systems. The Federal Power Commission assumes (Hydroelectric Power Evaluation, Report P-35, March 1968) that the annual load factor will be relatively high over the first half of the estimated 30- to 35-year service life of a power plant, then will diminish, giving a lifetime average of 55 to 60 percent. Dennis and Bernstein[3] have assumed a load factor of 60 percent over an assumed 11-year life for the emission control systems. Although their assumptions may be somewhat too conservative, the author believes that the average factor of 80 percent assumed in the present study is probably too optimistic.

Because of the high fixed charges associated with the emission control systems, the load factor exerts great leverage in determining the incremental cost of producing electricity. The estimates of incremental unit costs presented in Tables I and II are probably as low as they can be expected to be with respect to load factor. The actual costs would probably be higher.

2. Amortization Period

Depreciation for the emission control systems was based on an assumed useful life of 20 years. It was considered that any control system that might go into service at the present time will be obsolete within 20 years, if not in a shorter period. It was assumed that tax laws and the regulations of public utility commissions will permit the use of this period in determining taxes and rates for electrical power. At this time there appear to be no precedents to indicate how the existing depreciation schedules of the tax laws might be interpreted in some specific cases. Dennis and Bernstein[3] assumed a depreciation period of 11 years, as specified for chemical plants by the Internal Revenue Service. Their choice of depreciation period apparently was influenced by a conviction that the recovery processes would become obsolete within the 11-year period as well as by the IRS regulations.

Some components of sulfur dioxide recovery systems are relatively conventional items of chemical plant equipment. For example, a basically standard contact process plant would be used to convert concentrated sulfur dioxide from a recovery system to sulfuric acid. Allowances for depreciation of the acid plant might then depend upon whether the plant was owned by the power company and considered to be part of the utility, or owned by an adjacent chemical company and considered to be part of a chemical plant.

For the purposes of the present study, such considerations were ignored and it was assumed that all components of the control and recovery systems would be depreciated over the 20-year period. Nevertheless, it must be recognized that the assigned useful life of the

control system, like the load factor of the power plant, can exert a very great weight in determining the incremental unit cost of producing electricity. The author believes that, on the whole, the assumption of a useful life of 20 years is as optimistic as can now be made for any control system, and that the actual useful lives might well prove to be shorter. In such case, the incremental unit costs of producing electricity given in Tables I and III could be substantially increased.

3. Fixed Charges

The total fixed charge of 14.5 percent of the capital investment per year was adopted from a report by the Tennessee Valley Authority,[13] which was in turn adapted from guidelines suggested by the Federal Power Commission, and embodies the assumption of a 20-year depreciation period. Recent increases in both taxes and the cost of capital are producing rises in the fixed charges associated with electric power generation. The figure of 14.5 percent is therefore lower than may be appropriate at the present time. It has been used in the present study primarily for the sake of consistency with previous studies, and because the uncertainties in other factors assumed appear to make greater elaboration of the capital charge estimates of limited utility. The trend in taxes and capital costs will, however, tend to increase the incremental unit costs of electrical power generation estimated in Tables I and II.

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Appendix A

MODELS FOR HYPOTHETICAL POWER PLANTS

For the hypothetical power plant models, DPCE-NAPCA specified the following conditions:

1. The two model plants shall be an existing 500-MW plant located in central Pennsylvania and a new 1000-MW plant located in the Midwest on a large navigable river.

2. Both plants shall have dry-bottom boiler furnaces and shall use coal containing 3 percent of sulfur and 9 percent of ash. The flue gases shall contain 0.21 percent of sulfur oxides.

3. The base load shall be 7,000 hours per year (load factor 80 percent).

4. The existing plant shall have an economizer with an exit gas temperature of 750°F and an electrostatic precipitator operating at 325°F. The new plant shall have no temperature constraints on gas cleaning equipment. The height of the stack for the existing plant shall be 500 feet. Sufficient space shall be available to permit installation of any equipment needed.

5. The efficiency of removal of the sulfur dioxide shall be 90 percent. Regional air and water pollution regulations shall be applicable.

DPCE-NAPCA also supplied an analysis of the coal assumed to be burned and a calculated analysis of the flue gas formed:

Analysis of Coal, As Fired

Constituent	lb/100 lb, Coal as Fired
C (burned)	72.7
H_2	5.1
N_2	1.4
S	3.0
O_2	6.9
H_2O	1.0
Ash	9.0
	100.0

Flue Gas Analysis[1]

Component	Percent by Volume
CO_2	13.89
SO_2 and SO_3	0.2148
O_2	3.301
N_2	74.514
H_2O	8.08
	100.0

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1 Excess air used = 20 percent. Moisture content of air = 0.013 lb/lb dry air.

Quantity of Flue Gas Formed

Basis	Lb Moles Gas/100 lb Coal Burned
Wet	43.67
Dry	40.14

The remainder of the model conditions were derived by SRI from the foregoing conditions, supplemented by assumptions based on data in the literature, or on the experience or judgment of the author.

A. Conditions Common to Both Model Plants

1. Heating Value of Coal
   13,325 Btu/lb (calculated from DuLong Formula)

2. Flue Gas Generated
   Volume of gas per 100 lb of coal burned (wet basis)
   = 16,880 SCF (70°F, 1 atm)
   = 25,013 CF (325°F, 1 atm)
   = 43,335 CF (900°F, 1 atm)

3. Air Inleakage into Flue Gas Handling System
   Assumed nil

4. Fly Ash
   Ash = 9.0 lb/100 lb coal burned
   Unburned carbon = 0.35 lb/100 lb coal burned
   Total ash and unburned carbon = 9.35 lb/100 lb coal burned
   Fly ash (ash + carbon) leaving furnace in flue gas
   = 80 percent of total
   = 7.49 lb/100 lb coal burned
   Fly ash concentration in flue gas = 3.105 grains/SCF (70°F, 1 atm, wet basis)

5. Sulfur Oxides Concentrations
   $SO_2 + SO_3$ = 0.2148 percent by volume, wet basis
   SO_3 = 30 ppm = 0.003 percent by volume, wet basis
   SO_2 = 0.2118 percent by volume, wet basis

6. Sulfur Emission

All of sulfur assumed to go into flue gas

Total sulfur emitted = 3.0 lb/100 lb coal burned

Sulfur emitted as SO_3

= 0.0417 lb S/100 lb coal burned

= 0.1042 lb SO_3/100 lb coal burned

Sulfur emitted as SO_2

= 2.958 lb S/100 lb coal burned

= 5.916 lb SO_2/100 lb coal burned

7. Efficiency of Emission Control

Sulfur oxide removal efficiency = 90 percent

Allowable discharge of sulfur (as SO_2, SO_3, or H_2SO_4) not to exceed 10 percent of the input sulfur to the collection system, or 0.3 lb S/100 lb coal burned

8. Fuel Available for Reheating Flue Gas

No. 2 fuel oil (sulfur content 0.35 percent or lower)

9. Reductant Available for Reducing Sulfur Dioxide

Natural gas (heating value 1000 Btu/CF)

B. Model A Plant

1. Location

Altoona, Pennsylvania

2. Generating Capacity

500 MW

3. Temperature of Flue Gas

At exit from economizer: 750°F

At electrostatic precipitator: 325°F

4. Electrostatic Precipitator Efficiency

95 percent

5. Stack Height

   500 feet

6. Drives for Fans and Pumps

   Electric

7. Heat Rate

   10,000 Btu/kwh

8. Coal Consumption

   0.7505 lb/kwh

9. Coal Firing Rate

   375,250 lb/hr
   187.63 tons/hr

10. Flue Gas Flow Rate

    1,056,204 SCFM (70^{o}F, 1 atm, wet basis)
    1,564,355 CFM (325^{o}F, 1 atm, wet basis)

11. Fly Ash Leaving Furnace in Flue Gas

    28,106 lb/hr
    14.053 tons/hr

12. Fly Ash in Flue Gas Leaving Precipitator

    Concentration = 0.1553 grain/SCF (70^{o}F, 1 atm, wet basis)
    Mass emission rate = 1405.3 lb/hr

13. Sulfur Oxide Emissions

    SO_2 = 22,200 lb/hr
    = 11.100 tons/hr
    = 266.40 tons/day
    SO_3 = 391.01 lb/hr
    = 9384 lb/day

C. Model B Plant

1. Location

   Cairo, Illinois

2. Generating Capacity

   1000 MW

3. Gas Cleaning System

   No temperature limits apply. Unless some other gas cleaning system is specified as an integral part of the sulfur oxide control process, the power plant will be assumed to be equipped with an electrostatic precipitator operating on flue gas at a temperature of 325°F and having an efficiency of 99.5 percent.

4. Stack Height

   The plant will be assumed to have an 800-foot stack when not equipped with a sulfur dioxide control system, and a 500-foot stack when equipped with a sulfur dioxide control system of 90-percent efficiency.

5. Drives for Fans and Pumps

   Either electric or steam turbine

6. Heat Rate

   9600 Btu/kwh

7. Coal Consumption

   0.7204 lb/kwh

8. Coal Firing Rate

   720,400 lb/hr
   360.20 tons/hr

9. Flue Gas Flow Rate

2,027,690 SCFM(70°F, 1 atm, wet basis)

3,003,230 CFM (325°F, 1 atm, wet basis)

5,203,090 CFM (900°F, 1 atm, wet basis)

10. Fly Ash Leaving Furnace in Flue Gas

53,958 lb/hr

26.979 tons/hr

11. Fly Ash in Flue Gas Leaving Precipitator of 99.5 Percent Efficiency

Concentration = 0.01553 grain/SCF (70°F, 1 atm, wet basis)

Mass emission rate = 269.79 lb/hr

12. Sulfur Oxide Emissions

SO_2 = 42,619 lb/hr

= 21.309 tons/hr

= 511.43 tons/day

SO_3 = 750.66 lb/hr

= 18,016 lb/day

Appendix B

COST FACTORS USED IN MODEL STUDIES

In an actual case, the busbar cost of electricity and the cost of steam generated at a power plant are, of course, related to the cost of fuel and other plant operating costs as well as to the fixed costs, which are in turn related to the capital investment and the plant load factor. In practice, it appears that the complexity of cost accounting makes it difficult for even a utility company to assign costs precisely for an individual power plant in its system.

In the present study, reasonable costs were assumed for individual items such as fuel, steam, and electrical power. No attempt was made to reconcile these costs, but it was appreciated that the values adopted may actually not be entirely consistent with one another.

A. Fuel Costs

Coal

The cost of coal was estimated from data in "Steam-Electric Plant Factors (1967)," published by the National Coal Association. The costs (¢/million Btu) for coal as burned in Pennsylvania power plants were averaged for coals having heating values greater than 13,000 Btu/lb, which had higher unit costs than the lower-grade coals. The average cost per million Btu was applied to the hypothetical coal assumed in the present study (heating value 13,325 Btu/lb) to give the following costs:

31.2¢/million Btu
$8.30/ton

The average costs of coal used in Illinois, Indiana, Ohio, and Pennsylvania were lower, but applied to coals having heating values in the range of 10,000 to 12,000 Btu/lb. Presumably these coals had generally higher ash contents than the hypothetical coal.

Natural Gas

An estimate of the cost of natural gas was taken as the rounded average of the average costs of natural gas burned at power plants in Illinois, Indiana, Ohio, and Pennsylvania. The data were taken from "Steam-Electric Plant Factors (1967)." The value adopted was:

30¢/million Btu

The heating value of the gas was assumed to be 1,000 Btu/CF.

No. 2 Fuel Oil

The cost of No. 2 fuel oil in central Pennsylvania was estimated at SRI to be:

$4.25/barrel

B. Water

The costs of water for various uses were taken from lists of standard factors drawn from the literature or were figures used for estimating purposes at SRI.

	Cost
Raw makeup water	2¢/1000 gal
Once-through cooling water (temperature 80°F)	2¢/1000 gal
Process water	20¢/1000 gal
Boiler feed water	40¢/1000 gal
Retreatment of process steam condensates for boiler use	38¢/1000 gal

C. Steam

The cost of high-pressure steam was very roughly estimated at SRI as:

65¢/1000 lb (2400 psig, 1050°F)

The cost of the steam includes the cost of boiler feed water treatment.

It was assumed that low-pressure steam was available as needed, obtained by extraction from the power plant turbines. The value of low-pressure steam extracted from turbines, or exhausted from a backpressure turbine, was assumed to be proportional to the value of the high-pressure steam in the ratio of the enthalpies of the steam at the two conditions.

D. Electrical Power

The busbar cost of electrical power generated and used in the power plant itself (including the emission control system), after inspection of data published by the Federal Power Commission and after discussions with staff of the Pacific Gas & Electric Co., was assumed to be as follows:

Generating cost	3.2 mills/kwh
Fixed charges	2.3 mills/kwh
Total cost	5.5 mills/kwh

E. Salaries and Payroll Benefits

Salaries

Operating labor	$3.75/hour
Supervision	$4.75/hour

Payroll benefits

25 percent of labor and supervision

F. Maintenance

Taken as an appropriate percentage of capital investment to cover both labor and materials.

G. Indirect Costs

Overhead

50 percent of labor, supervision, and maintenance

H. Fixed Charges

The total fixed charge was adopted from the Tennessee Valley Authority's Conceptual Design and Cost Study No. 2 (1969) for National Air Pollution Control Administration, "Sulfur Oxide Removal from Power Plant Stack Gas--Use of Limestone in Wet-Scrubbing Process." The breakdown of fixed charges (adapted from Federal Power Commission guidelines) is given as follows:

	Annual Percent of Capital Investment
Depreciation (20-year life)	5.00
Insurance	0.25
Cost of capital (capital structure assumed to be 50% debt and 50% equity)	
Bonds at 6% of average undepreciated investment	1.50
Equity at 11% of average undepreciated investment	2.75
Taxes	
Federal	2.80
State	2.20
Total fixed charges	14.50

Appendix C

STACKS FOR USE ON CONTROLLED SULFUR DIOXIDE EMISSION SOURCES

Evidently, no clear policy has yet been developed with respect to choice of the heights of stacks to be used on power plants equipped with sulfur dioxide emission controls. The use of a high stack on an uncontrolled emission source reduces the maximum ground-level concentration of sulfur dioxide at the point where the plume reaches the ground, but it does not, of course, reduce the total emission of sulfur dioxide that may affect areas beyond the range of influence of the stack. If emission controls are used as well, not only is the maximum ground-level concentration further reduced, but the quantity of sulfur dioxide affecting the area beyond the range of influence of the stack is reduced also.

If, on the other hand, the use of emission controls is accompanied by a reduction of the stack height, the reduction of the maximum ground-level concentration of sulfur dioxide may be partly lost, depending upon the balance between the reduction in stack height and the reduction in emission. The point of maximum ground-level concentration will be brought closer to the emission source.

There is reluctance to make major reductions in stack height, even with installation of sulfur dioxide emission controls, because of concern with other contaminants that may still remain in the flue gas and with the possibility of failures of the emission control system.[C4]

If the sulfur dioxide emission control system reduces the temperature of the flue gas, the effect is to reduce the effective stack height because the density of the gas is increased and hence its buoyancy is reduced.[C3,C4] When the hot gas is scrubbed, the effect of gas cooling is counteracted to some degree by the lowering of the gas density due to the increase in the water vapor content of the gas stream.

Particular concern has been expressed about "negative buoyancy." If the plume emerging from the stack is saturated and contains water droplets, the subsequent evaporation of the drops in the atmosphere may cool the gas sufficiently to cause the plume to descend rapidly.[C1,C2] The presence of the droplets in the plume may directly produce such a descent by increasing the effective density of the gas in the plume. However, the author believes that the presence in stack gas of any such relatively gross quantities of liquid would most likely be due to entrainment from the scrubber, which can be prevented. It has been suggested[C1] that at least some of the reported instances of rapid descent of plumes of scrubbed gas may have resulted from downwash produced by adjacent structures. Plumes of scrubbed gas that is saturated but still substantially above ambient temperature have been commonly observed to rise and disperse in the same general manner as do warm, dry plumes. Careful observations of such plume behavior have been reported by Boyer et al.[C1]

The necessity for reheating scrubbed stack gases to ensure adequate dispersion is not entirely clear, at least if the efficiency of the scrubber is high and some minimum stack height is used. Using a hypothetical 200-MW power plant unit with a 300-foot stack as an example, TVA[C4] calculated the effects of cooling (by scrubbing) and of reheating the flue gas on the maximum ground-level concentration of sulfur dioxide, assuming varying scrubber efficiencies. The results indicated that if the scrubber had an efficiency of at least 90 percent and the stack height was at least 300 feet, failure to reheat the gas would have only a small effect upon the percentage reduction in the maximum ground-level concentration of sulfur dioxide achieved by scrubbing. Applying a commonly used rule of thumb that a stack should be at least 2.5 times the height of adjacent structures, a typical power plant should in any case use a stack at least 300 feet in height in order to avoid possible downwash of the gas.

In the present study, it was assumed that in the Model A plant, the treated flue gas would be discharged through the existing 500-foot stack.

In the case of the Model B plant, it was assumed that the plant would be equipped with an 800-foot stack if it had no sulfur dioxide emission controls, and with a 500-foot stack if it were equipped with an emission control system of 90 percent efficiency. Such a reduction in stack height does not appear to be unreasonable. In the light of present knowledge, a minimum stack height of 500 feet for the controlled emission source seems appropriate.

The Model B control systems (both Cat-Ox and Wellman-Lord) were each given credit for the differential between the costs of the 800- and 500-foot stacks. A rough estimate of the cost differential for stacks 30 feet in inside diameter ($850,000) was made from Fig. 6-1 of "Control Techniques for Sulfur Oxide Air Pollutants."[C2]

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Appendix D

ESTIMATION OF THE VALUES OF SULFUR BY-PRODUCTS*

As applied to power plant flue gases, the Monsanto Cat-Ox system can produce only 78-percent sulfuric acid. The Wellman-Lord system produces concentrated sulfur dioxide, which has relatively insignificant markets as such but can be converted to elemental sulfur or to either 93-percent or 98-percent sulfuric acid.

The estimates of the sulfur by-product prices at the two hypothetical power plant sites (Altoona, Pennsylvania for the Model A plant, and Cairo, Illinois for the Model B plant) were based on an assumed Gulf Coast price for sulfur of $30/long ton. This Gulf Coast price was estimated during the previous study[D1] as a likely average over the period to 1975. Currently, sulfur prices have become chaotic as sulfur supplies have shifted abruptly from shortages to surpluses. Apparently it is scarcely possible to define an "established" price for sulfur. The development of sulfur surpluses has followed a continuing period of low activity in the market for fertilizers, which provides the largest outlet for sulfur.

In the next five years, the withdrawal of marginal producers and a possible revival of the fertilizer market may cause sulfur prices to rise from their present lows. However, the figure of $30/long ton now appears to be an optimistic one. It perhaps represents the upper end of the probable range of Gulf Coast prices to be encountered in the period to 1975. It cannot even be assumed that the Gulf Coast sulfur price will continue to maintain its previous status as the base line for world sulfur prices. Nevertheless, if the above limitations are recognized, the assumption of the $30/long ton price is probably as good as any that can be made at this time. Prices at other locations can be estimated by adding to the Gulf Coast price the cost of transportation from the Gulf

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*This material was prepared by F. Alan Ferguson, Industrial Economist, SRI.

Coast. This procedure assumes, of course, that the incursion of the sulfur from a new source will not be large enough to disrupt the local markets.

Model A Plant - Altoona, Pennsylvania

The closest and most ready markets for sulfur by-products produced at the hypothetical power plant Model A would be located in the Pittsburgh area, about 120 miles away. The acid users in the Pittsburgh region currently consume over 500,000 tons of acid per year, a large portion of which is used to clean (pickle) iron and steel products and to produce ammonium sulfate from the ammonia in coke oven gas. Both of these uses could employ 78-percent sulfuric acid as well as 93-percent or 98-percent acid, and the total market is more than large enough to absorb all the acid that could be produced at the Altoona plant. The Pittsburgh area could also absorb elemental sulfur that might alternatively be produced at Altoona.

The prices that could be obtained for the various possible by-products producible at Altoona will depend upon a number of factors, a few of which can be allowed for at the present time. On the basis of a Gulf Coast sulfur price of $30/long ton, the delivered price of elemental sulfur in the Pittsburgh area will be about $36.50/long ton. Since the elemental sulfur now being sold in the Pittsburgh area is merchant sulfur, that produced at Altoona could probably capture about half the existing market if sold at the same price of $36.50/long ton. Since the cost of transportation from Altoona to Pittsburgh would be about $3.50/long ton, the f.o.b. price of the sulfur at Altoona might be $33/long ton.

Estimating the price of Altoona sulfuric acid is more difficult because some of the Pittsburgh area acid is made for captive consumption and the rest is made and sold as merchant acid. The prices of captive and merchant acid can differ by as much as a factor of two. The price of merchant acid will vary according to whether the acid is sold on long term contracts or when and as needed (spot sales). However, relatively little acid is sold at the higher spot prices. The highest price the plant at Altoona could expect to receive for its acid at Pittsburgh

would be the same as that which would return a reasonable profit to an acid producer operating a contact acid plant and using Gulf Coast sulfur as raw material. The cost of producing acid from sulfur costing $36.50/ long ton would be about $13/short ton (100 percent acid basis). With the allowance of about 15 percent as a return on investment, the corresponding price of acid would be $15/short ton (100 percent acid basis). The corresponding f.o.b. prices at Altoona can be estimated by subtracting the transportation charges of $4/ton (100 percent acid basis) for 93-percent acid and $5/ton (100 percent acid basis) for 78-percent acid giving:

93-percent acid: $12/ton of 100 percent acid

78-percent acid: $11/ton of 100 percent acid

If it should be necessary to displace existing captive producers to capture adequate markets, the Altoona acid might have to sell at Pittsburgh for about $2/ton less than the cost of production from sulfur, or $11/ton. The corresponding f.o.b. prices at Altoona would then be:

93-percent acid: $7/ton of 100-percent acid

78-percent acid: $6/ton of 100-percent acid

A detailed market analysis would be required to determine whether the price of the Altoona acid would be nearer the upper end ($11 to $12/ton) or the lower end ($6 to $7/ton) of the price range estimated.

Model B Plant - Cairo, Illinois

Finding markets for elemental sulfur produced at a plant at Cairo should not be difficult. In 1965, over 1.2 million tons of elemental sulfur were barged along the Ohio River and the upper Mississippi River. Assuming a Gulf Coast sulfur price of $30/long ton, sulfur produced along the rivers should sell for $30 plus the cost of barging sulfur from the Gulf Coast to the actual plant site. The cost of barging elemental sulfur from the Gulf to Cairo, Illinois is about $2.90/long ton. Hence, sulfur produced at the Cairo plant should sell for about $33/long ton.

If sulfuric acid, rather than elemental sulfur, should be produced, finding markets for large quantities of acid might be substantially more difficult. In 1966, only about 130,000 tons of acid were barged along the Ohio River and 30,000 tons along the upper Mississippi River. From the summary (Table D-1) of uses of sulfuric acid near three sites on these two rivers, it is apparent that the only end use large enough to consume the quantities of acid that could be supplied from the Cairo plant would be the production of phosphate fertilizers.

A recovery system producing sulfuric acid at Cairo might supply either existing phosphate fertilizer manufacturers or a new plant located at or near Cairo to take advantage of the local source of acid. Supplying such a new fertilizer manufacturer would probably yield the highest price for the acid, although a more detailed market analysis would be required to determine whether this would actually be the case. Nevertheless, the sulfuric acid would have to be priced to provide the new fertilizer producer with a competitive advantage over the existing suppliers. The existing producers of phosphate fertilizers that are capable of supplying fertilizer to the midwestern market at the lowest prices are probably those who manufacture their product at plant sites in Louisiana and ship it to the markets by barge. The price of the by-product acid at Cairo must therefore permit the new producer to make his fertilizer for less than the Louisiana producers can make and ship theirs.

In Table D-2 a comparison is made of the costs of producing 54-percent wet process phosphoric acid in Louisiana and at the hypothetical Cairo site. The data indicate that the Cairo producer could pay about $32.20 for the amount of 100-percent sulfuric acid necessary to make 1.85 tons of 54-percent (P_2O_5) phosphoric acid, which is equivalent to 1.0 ton of P_2O_5. Since 2.7 tons of sulfuric acid (100-percent basis) are required to produce the 1.85 tons of phosphoric acid, the hypothetical producer could pay $11.94/ton (100-percent acid basis) for the sulfuric acid and be able to make phosphoric acid to sell for the same price as that made in Louisiana and shipped to the Cairo area. However, the sulfuric acid would probably have to be offered at somewhat less than $11.94/ton in order to provide incentive to the new phosphoric acid producer. Under

these circumstances, the sulfuric acid could probably be sold for $10 to $11/ton (100-percent acid basis). Since little or no transportation would be required, 78-percent and 93-percent acid should sell for the same price per unit quantity of 100-percent acid.

REFERENCES

D1. McKee & Company, Arthur G., Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry, Final Report to National Air Pollution Control Administration, June 1969, Contract No. PH 86-68-85.

Table D-1

ESTIMATED SULFURIC ACID DEMAND IN SELECTED PRODUCING AREAS, 1966
(thousands of short tons 100% H_2SO_4)

	Cairo-Paducah	St. Louis	Chicago
Phosphate fertilizers	107	789	1,375
Ammonium sulfate	13	5	113
Iron and steel pickling	--	3	169
Rayon and cellulose films and fibers	--	--	74
Petroleum refineries	10	81	78
Sulfonated detergents	15	57	33
Ammonium sulfate	--	28	12
Titanium dioxide	--	410	--
Hydrofluoric acid	45	--	38
Total Major End Uses	190	1,373	1,892

Source: A. D. Little, Inc.

Table D-2

PHOSPHORIC ACID PRODUCTION COSTS
(per ton 100% P_2O_5)

	Plant Location			
	Louisiana		Cairo, Illinois	
	Quantity Required	Cost/Ton 100% P_2O_5	Quantity Required	Cost/Ton 100% P_2O_5
Phosphate rock (72 bpl)				
$6.00/ton + $3.25 frt	3.2 tons	$29.60		
$6.00/ton + $5.85 frt			3.2 tons	$37.92
Elemental sulfur				
$30/long ton + $1.50 frt	0.82	25.83		
			0	0
H_2SO_4 manufacture $3.40 per ton (excluding cost of sulfur but including 15% return on investment)	2.7	9.18	0	0
Phosphoric acid manufacture (excluding raw material cost and profit)		14.00		14.00
Barge freight $3.00/ton (phosphoric acid from Louisiana to Cairo, Ill.)	1.85	5.55		0
Subtotal		$84.16		$51.92
H_2SO_4 purchase price $11.94/ton (to match cost of producing phosphoric acid at Cairo, Ill. and in Louisiana)		0	2.7	32.24
Total		$84.16		$84.16

Source: SRI